

DESIGN OF BINARY VAPOR CONDENSERS
USING THE COLBURN-DREW EQUATIONS

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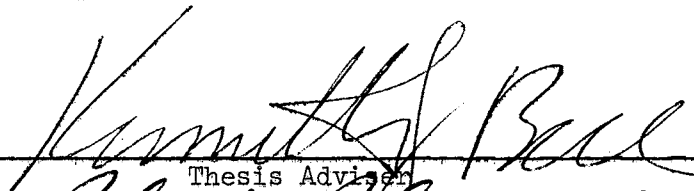
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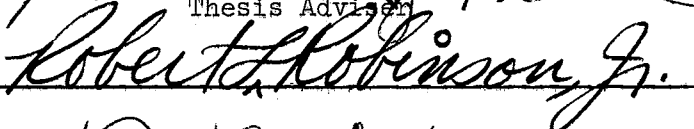
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PREFACE

The purpose of this study is to present a computer program for the design of binary vapor condensers using the basic heat and mass transfer equations formulated by Colburn and Drew in 1937.

In order to make these equations useful in a design sense they must be integrated over the entire condensation process. This may be accomplished by the use of a forward difference method whereby the condenser is broken up into area increments and the basic transport equations along with the appropriate heat and material balance are applied in a stepwise manner. The results of this method are compared with those of a commonly used approximation method for two different cases.

I wish to express my sincere thanks to Dr. Kenneth J. Bell for his advice and guidance and to the Phillips Petroleum Company for their fellowship grant which made this work possible. Also, I wish to thank Dr. John H. Erbar for his assistance in obtaining thermodynamic and physical property data necessary to perform the desired calculations.

I wish to express my sincere gratitude to my wife, Brenda, for her help and moral support during this research project.

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NOMENCLATURE

ΔA	- Area increment, ft^2
b	- Diffusion coefficient, ft^2/hr
C_o	- Quantity defined by eqn. (3-7), dimensionless
C_{pA}, C_{pB}	- Heat capacity of component A, component B, vapor (mean), Liquid, coolant; $\text{Btu}/\text{lb } ^\circ\text{F}$
C_{pM}, C_{pL}	
C_{pc}	
D	- Tube diameter, $\text{ft}.$, D_i - inside diameter, D_o - outside diameter
F	- Quantity defined as kgP , where kg is defined in reference (2) as $bP/RTZP_{BM}$.
G	- Mass velocity of the total flow, $\text{lb}/\text{ft}^2 \text{ hr}$
g	- Gravitational acceleration, ft/sec^2
h_G, h_{TP}	- Heat transfer coefficient for the vapor phase, vapor phase with two-phase enhancement, condensation, liquid phase, fouling inside the tube, fouling outside the tube, coolant; $\text{Btu}/\text{ft}^2 \text{ M}^\circ\text{F}$
h_c, h_l	
h_{fi}, h_{fo}	
h_o	
j	- Mass transfer or heat transfer factor
k	- Thermal conductivity, $\text{Btu}/\text{ft hr}^\circ\text{F}$
\dot{M}_v, \dot{M}_l	- Mass flow rate of the vapor, the liquid and coolant; lb/hr
\dot{M}_c	
M, M_A	- Mean molecular weight, molecular weight of Component A, molecular weight of Component B; $\text{lb}/\text{lb mole}$
M_B	

- N_A, N_B - Mass fluxes of A and B, $\text{lb/ft}^2 \text{ hr}$
 Pr_L - Liquid phase Prandtl number
 $(Q/A)_S, (Q/A)_L$ - Sensible heat flux, latent heat flux, total heat flux
 $(Q/A)_T$ $\text{Btu/ft}^2 \text{ hr}$
 q_{sv}, q_{sL} } - Sensible vapor heat load, sensible liquid heat load,
 q_λ, q_{TOT} } latent heat load, total heat load; Btu/hr
 R - "Two phase" enhancement factor (h_{TP}/h_G) defined in
eqn (3-13), dimensionless
 Re_c - Reynolds number for condensing coefficient calculation
 r_w - Wall resistance, $(\text{Btu/ft}^2 \text{ hr}^\circ\text{F})$
 T_v, T_i - Vapor bulk temperature, interfacial temperature, cool-
 T_c ant temperature $^\circ\text{F}$
 U - Overall heat transfer coefficient, $\text{Btu/ft}^2 \text{ hr}^\circ\text{F}$
 y_i, y_v - Mole fraction of more volatile component at the inter-
face and in the bulk vapor
 x - Quality (lb vapor/lb total)
 $x_{l,i}$ - Liquid mole fraction of more volatile component
 z $N_A / (N_A + N_B)$

Greek Letters

- η - Fractional distance through the film ($\frac{\text{distance into film}}{\text{film thickness}}$)
 $X_{tt}, \phi_{\lambda tt}$ - Martinelli-Nelson factors defined by eqns. 3-17 and 18
 λ - Latent heat of vaporization, Btu/lb
 μ - Viscosity; lb/ft hr
 ρ - Density; lb/ft^3

Subscripts

- v - Vapor phase
- l - Liquid phase
- f - Friction
- TPF - Two Phase flow

CHAPTER I

INTRODUCTION

The condensation of a binary vapor differs in several aspects from the condensation of a pure vapor. The process in general is non-isothermal. Because the components of the mixture will have different boiling points at the system pressure, the heavier component will be preferentially condensed. This preferential condensation gives rise to a decrease in the dew point of the remaining vapor and thus the vapor temperature decreases continuously over the condensation process. In addition, the condensate already laid down and the remaining vapor must be cooled further. All these sensible heat effects, while generally being small compared to the latent heat load, must be taken into account in the condenser design. The sensible heat removed from the vapor phase is of particular importance because of the relatively poor heat transfer rate associated with this process.

Another essential difference between binary and pure component condensation is the resistance to heat and mass transfer in the vapor phase for the binary case. A diagram of the condensation process at a point in the condenser is shown in Figure 1. This diagram shows a vapor stream at temperature T_v and bulk composition y_v flowing past a tube wall. The vapor is separated from the wall by a layer of condensate. It is assumed that at the vapor-liquid interface the phases are in thermodynamic equilibrium. Thus the compositions x_{1_i} and y_{1_i}

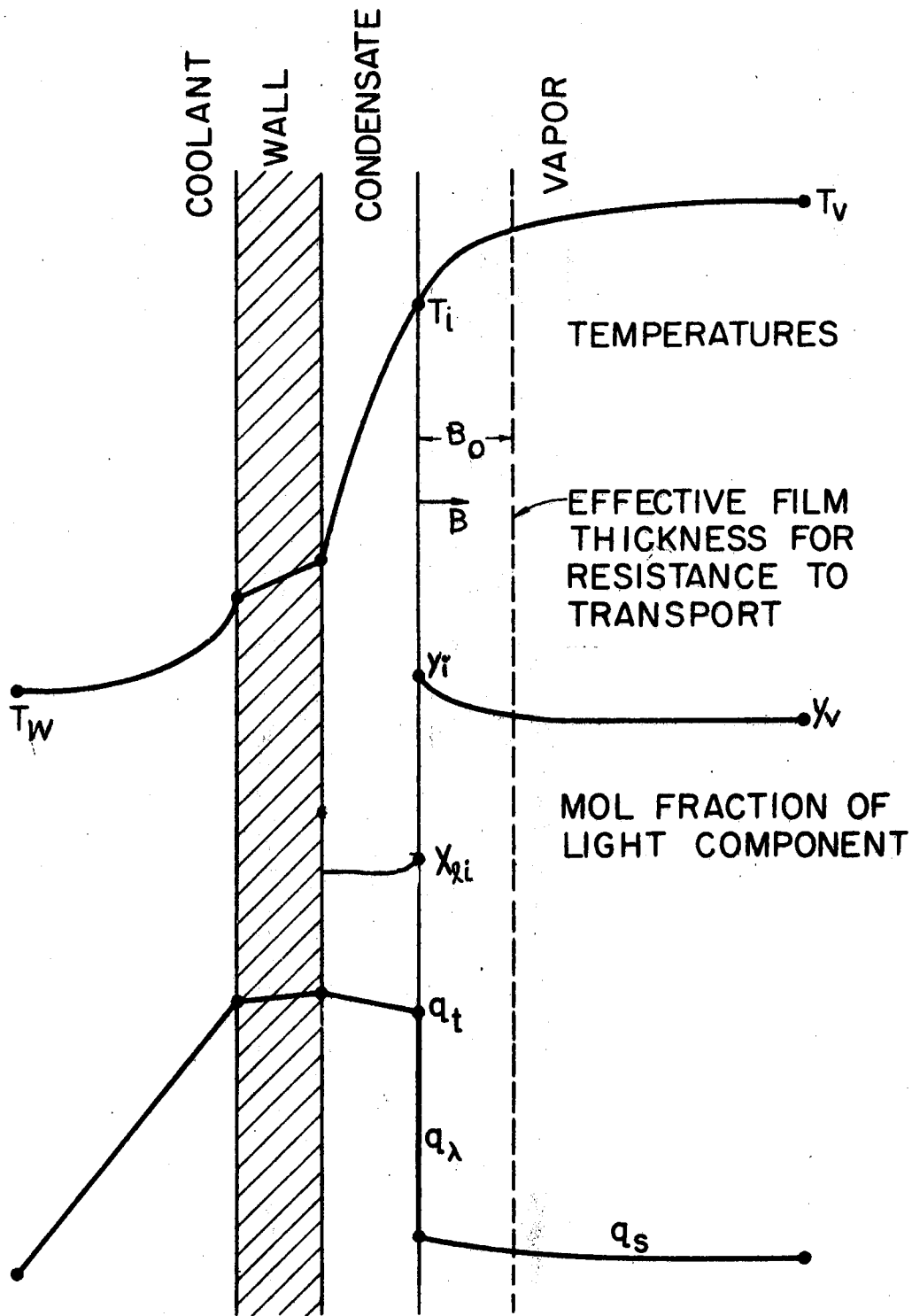


Figure 1. Diagram of the Condensation Process

are the equilibrium compositions at T_i and the system pressure. In general, for a finite condensing rate y_v and y_i will not be equal; further, since T_i is less than T_v , y_i will be greater than y_v . This analysis indicates that the gross mass flow of the vapor to the condensing interface carries with it a somewhat higher concentration of the more volatile component.

The non-uniformity of the vapor phase and the fact that as the heavier component is preferentially condensed the compositions change from point to point indicate that there are diffusional resistances in both phases that are coupled with the heat transfer processes. The change in the composition suggests that the transport equations must be applied in a point-wise manner.

Since the bulk vapor and liquid phases are not in equilibrium, the familiar "inverse lever rule" cannot be applied to determine the fraction of the inlet vapor stream that has condensed at any point in the condenser. In order to calculate the condensing rate, phase compositions, and the total condensate flow rate at any point, it is necessary to carry out an incremental mass balance for each phase.

It is necessary to carry out an incremental energy balance in order to calculate the vapor temperature profile. The energy balance calculation must take into account both the sensible and latent heat loads as was previously mentioned. These heat effects are directly related to the mass transfer rate; thus, all the computations are coupled and reiterative making use of a computer imperative to solve this set of equations.

The purpose of this study is to present a computer program which solves all the necessary equations for any set of specified conditions.

The cases presented are for binary condensable vapors having miscible condensates, using a countercurrent coolant of constant specific heat. Certain assumptions are made in the solution of the system equations. First, the thermodynamic and physical properties of both the liquid and vapor phases are assumed to be known. Second, the condensation process is assumed to be isobaric. Third, the condensate film is assumed to be well mixed. This implies that there is no diffusional mass transfer resistance in the liquid phase.

The Colburn-Drew design method is applied to two different systems and the results are compared to an approximate method applied to the same cases.

CHAPTER II

LITERATURE SURVEY

The basic transport equations describing binary condensation were first presented by Colburn and Drew (1). A somewhat more modern presentation of the same work is given by Sherwood and Pigford (2). Although the work by Colburn and Drew provided the transport equations necessary to perform the calculations for a condensing binary vapor, the authors made no attempt to apply these equations along with the necessary heat and material balances to the actual design of a condenser of finite area. In addition, there seems to have been little further attempt to put the Colburn-Drew formulation to work in this sense. Most of the work in the area of design has been aimed at approximation methods.

Kent and Pigford(3) studied experimentally and theoretically the performance of a reflux condenser for a toluene-ethylene dichloride system. Their results showed mass transfer effects in agreement with the Colburn-Drew theory. Several other experimental studies (4,5,6) have been carried out which tend to support the Colburn-Drew model, but they have not extended the model to a design mode.

One type of binary condensation which has been solved by a rigorous method is the condensation of a pure component from an insoluble gas. Colburn and Hougen (7) presented the equations and design procedure. This method is quite similar to the Colburn-Drew method for binary

condensation except in the latter case the condensate is not a pure component. This introduces another loop into the reiterative solution of the system equations. The Colburn-Drew equations as presented in this paper could be used for the non-condensable case with the proper modifications.

Several general purpose design methods have been proposed. Kern (8) recommended ignoring the vapor phase resistances to heat and mass transfer and increasing the area computer by condensation alone by the ratio of the total heat load to the latent heat load. Kern also indicated how to use the non-linear condensing temperature profile instead of the LMTD.

Gully (9) attacked the problem from the standpoint of computing an adjusted mean temperature difference. Gloyer (10) proposed a method which also neglects the vapor phase resistances and is based on design using terminal conditions of the condenser. Gloyer concentrates on correcting the MTD and the vapor phase coefficient with various weighting factors.

An approximate design procedure was published by Ward (11) and later extended into a more meaningful form by Bell and Ghaly (12). The latter method takes into account the sensible heat load due to liquid phase cooling. The essential assumption in both methods is that the mass transfer resistance in the vapor may be neglected and compensated for by overestimating the resistance to sensible heat transfer. Another assumption made is that the bulk vapor and condensate streams are in thermodynamic equilibrium at any point in the process.

The design equation used in the Bell-Ghaly method for the condenser heat transfer area is:

$$A = \int_0^{Q_T^*} \frac{1 + \frac{ZU_o}{hsv}}{U_o (T_v - T_w)} dQ \quad (2-1)$$

where Z is the ratio of the local vapor sensible heat flux to the local total heat flux:

$$Z = (Q/A)_S / (Q/A)_T \quad (2-2)$$

and U_o is the overall coefficient for the condensate film, dirt, wall, and coolant:

$$U_o = \frac{1}{\frac{dA_o}{h_i dA_i} + \frac{dA_o}{h_{fi} dA_i} + \frac{\Delta X_w dA_o}{k_w dA_i} + \frac{1}{h_{fo}} + \frac{1}{h_c}} \quad (2-3)$$

The quantity (ZU_o/hsv) is a measure of the additional area required to remove vapor sensible heat. The required quantities for equation 2-1 are calculated for each point in the condenser and the area required is then found by a graphical integration of Equation 2-1.

CHAPTER III

THE BASIC EQUATIONS

Vapor Phase Transport Equations

The Colburn-Drew formulation of the transport equations is based on two main assumptions. First, it is assumed that all the vapor phase resistance to heat and mass transfer is in a laminar boundary layer or film of thickness B_0 adjacent to the vapor-liquid interface. Second, it is assumed that, at the vapor-liquid interface, the compositions of the phases are the equilibrium compositions at the interfacial temperature and system pressure.

The net molar flux of the more volatile component A to the interface, N_A , arises from two sources. First, the contribution due to the molecular diffusion of the more volatile component relative to the heavier is:

$$N_A = F \frac{dy}{d\eta} \quad (3.1)$$

where η is the fractional distance through the film B/B_0 measured from the interface out into the vapor as shown in Figure 1. The second source is the flux of Component A due to bulk vapor flow which is merely the total flux $(N_A + N_B)$ times the mole fraction of Component A in the vapor phase.

The next molar flux of Component A is merely the sum of these two contributions:

$$N_A = (N_A + N_B)y + F \frac{dy}{d\eta} \quad (3-2)$$

If the composition of the condensing vapor is z , then:

$$N_A = (N_A + N_B)z \quad (3-3)$$

Equation 3-2 then becomes

$$(N_A + N_B)d\eta = F \frac{dy}{z - y} \quad (3-4)$$

Integration of this equation between the limits of $\eta = 0, y = y_i$ and $\eta = 1, y = y_v$ gives:

$$(N_A + N_B) = F \ln \left[\frac{z - y_i}{z - y_v} \right] \quad (3-5)$$

Similarly, the sensible heat flux from the bulk to the interface is the sum of the heat flux due to the bulk flow of the vapor to the interface and the heat conducted through the laminar film:

$$\left(\frac{Q}{A}\right)_s = (N_A M_A C_{PA} + N_B M_B C_{PB})(T_s - T_i) + h_g \frac{dT}{d\eta} \quad (3-6)$$

The first term is the flux arising from the bulk flow; the second term is the flux due to conduction. Integration of this equation across the boundary layer with the limits: $\eta = 0, T = T_i$ and $\eta = 1, T = T_v$ gives:

$$\left(\frac{Q}{A}\right)_s = \frac{Co}{1 - e^{-Co}} \left[h_g (T_v - T_i) \right] \quad (3-7)$$

where:

$$C_o = \frac{1}{h_G} \left[N_A M_A C_{PA} + N_B M_B C_{PB} \right]$$

The latent heat flux resulting from the condensation of $(N_A + N_B)$ moles of vapor is:

$$\left(\frac{Q}{A}\right)_L = (N_A + N_B)M\lambda = F \ln \left[\frac{z - y_i}{z - y_v} \right] M\lambda \quad (3-8)$$

The total heat flux is just the sum of the latent and sensible fluxes:

$$\left(\frac{Q}{A}\right)_T = \frac{C_o}{1 - e^{-C_o}} \left[h_G (T_v - T_i) \right] + F \ln \left[\frac{z - y_i}{z - y_v} \right] M\lambda \quad (3-9)$$

This total flux must be equal to the flux from the vapor liquid interface to the coolant (neglecting sensible cooling of the condensate):

$$\left(\frac{Q}{A}\right)_T = U(T_i - T_c) \quad (3-10)$$

where U is the overall heat transfer coefficient from the vapor-liquid interface to the coolant.

Calculation of the Vapor Phase Heat and Mass Transfer Coefficients

The vapor phase heat transfer coefficients were calculated in two different ways: 1) assuming that the heat and mass transfer are for the vapor phase only flowing on a smooth conduit, and 2) assuming that

the presence of the liquid phase enhances the heat and mass transfer coefficients by disturbing the boundary layer.

The single phase heat transfer coefficient was calculated using a correlation by Chilton and Colburn (13)

$$h_G = j_H C_{PM} G_x \left(\frac{C_P \mu}{k} \right)^{-2/3} \quad (3-11)$$

where:

$$j_H = .023 \left(\frac{DG_x}{\mu} \right)^{-.2} \quad (3-12)$$

The "two phase" coefficient correlation used was developed from the relationship between the heat transfer coefficient and pressure gradient for single phase flow:

$$h_G = K \left(\frac{dp}{dl} \right)_{f,v}^{.445} \quad (3-13)$$

where K is a constant which combines the constants (physical properties, etc.) appearing in the equations for the heat transfer coefficient and pressure gradient. By analogy:

$$h_{TP} = K \left(\frac{dp}{dl} \right)_{f,TPF}^{.445} \quad (3-14)$$

then:

$$\frac{h_{TP}}{h_G} = \left[\frac{\left(\frac{dp}{dl} \right)_{f,TPF}}{\left(\frac{dp}{dl} \right)_{f,v}} \right]^{.445} \quad (3-15)$$

The frictional pressure gradient for two phase flow is calculated by the Martinelli-Nelson correlation (15):

$$\left(\frac{dp}{dl}\right)_{f,TPF} = \phi_{l_{tt}}^2 X_{tt}^{1.75} \left(\frac{dp}{dl}\right)_{f,v} \quad (3-16)$$

where:

$$X_{tt} = \frac{(1-x)}{x} \left(\frac{\rho_v}{\rho_l}\right)^{.57} \left(\frac{\mu_l}{\mu_v}\right)^{.11} \quad (3-17)$$

and:

$$\phi_{l_{tt}}^2 = \left[1 + \frac{1}{X_{tt}} + \frac{3(1-p/p_c)}{\sqrt{X_{tt}}} \right]^{1.75} \quad (3-18)$$

The vapor phase mass transfer coefficient used was calculated by the equation:

$$F = j_M \left(\frac{Gx}{M}\right) \left(\frac{\mu}{\rho b}\right)^{-2/3} \quad (3-19)$$

where $j_M = j_h$

The "two phase" mass transfer coefficient used was calculated in the same manner as the "two phase" heat transfer coefficient:

$$\frac{F_{TP}}{F} = \left[\frac{\left(\frac{dp}{dl}\right)_{f,TPF}}{\left(\frac{dp}{dl}\right)_{f,v}} \right]^{.445} \quad (3-20)$$

where the pressure gradients are the same as in equation 3-15.

Calculation of the Condensate Film Coefficient

Inside Vertical Tubes

For the calculation of the condensing coefficient in vertical tubes there are three cases which must be considered. For the case of low tube loadings and condensate flow rates, the condensate film flow is gravity-dominated and laminar, and the Nusselt equation (16) is used:

$$h_c = 1.47 \left[\frac{k_\ell^3 \rho_\ell (\rho_\ell - \rho_v) g}{\mu_\ell^2} \right]^{1/3} \text{Rec}^{-1/3} \quad (3-21)$$

At higher condensate rates the film becomes turbulent and the Colburn correlation (17) is the correct one to use. A curve-fitted form of the Colburn correlation is:

$$h_c = .011 \left[\frac{k_\ell^3 \rho_\ell (\rho_\ell - \rho_v) g}{\mu_\ell^2} \right]^{1/3} \text{Rec}^{1/3} \sqrt[4]{\text{PR}_\ell} \quad (3-22)$$

where

$$\text{Rec} = \frac{DG(1-x)}{\mu} \quad \text{and} \quad \text{PR}_\ell = \left(\frac{Cp\mu}{k} \right)_\ell \quad (3-23)$$

At high tube loadings the condensate film is vapor-shear controlled and the Boyko-Kruzhilin correlation (18) is valid:

$$h_c = .024 \frac{k_\ell}{D} \left(\frac{DG}{\mu_\ell} \right)^{.8} \left(\text{PR}_\ell \right)^{.43} \left[\frac{\sqrt{(\rho/\rho m)_i} + \sqrt{(\rho/\rho m)_o}}{2} \right] \quad (3-24)$$

where

$$\left(\frac{\rho}{\rho_m}\right)_i = 1 + \left(\frac{\rho_l - \rho_v}{\rho_v}\right) x_i \quad (3-25)$$

and

$$\left(\frac{\rho}{\rho_m}\right)_o = 1 + \left(\frac{\rho_l - \rho_v}{\rho_v}\right) x_o \quad (3-26)$$

The procedure is to calculate h_c by all three equations and choose the higher value, since that is the one that is valid under existing conditions.

Inside Horizontal Tubes

For condensation inside horizontal tubes a procedure similar to that suggested for vertical tubes is used. At low tube loadings (gravity dominates) the Kern correlation (8) is valid:

$$h_c = .761 \left[\frac{k_l^3 \rho_l (\rho_l - \rho_v) g L}{\mu_l W_T} \right]^{1/3} \quad (3-27)$$

At high condensing loads, with vapor shear dominating, the Boyko-Kruzhilim correlation (Equation 3-21) is valid. The procedure again is to calculate h_c from both equations and choose the higher value.

Calculation of the Overall Heat

Transfer Coefficient

The overall heat transfer coefficient from the vapor liquid interface to the coolant is:

$$U_i = 1 / \left(\frac{1}{h_c} + \frac{1}{h_{fi}} + \frac{D_i X_w}{D_L k_W} + \frac{D_i}{D_o h_{fo}} + \frac{D_i}{D_o h_o} \right) \quad (3-28)$$

The values of the coolant heat transfer coefficient, h_o , the fouling coefficients h_{fi} and h_{fo} , the the wall dimensions, and thermal conducting are assumed to be constant and are supplied as input data.

Mass and Energy Balances

In order to apply the Colburn-Drew equations to a condenser design, the condenser is broken up into area n increments, as shown schematically in Figure 2. For each area increment, component and overall mass balances are written for each phase. For the vapor phase:

Light component:

$$\dot{M}_{v_{i-1}} y_{v_{i-1}} = \left[(N_A + N_B) zM \right] \Delta A + \dot{M}_{vi} y_{vi} \quad (3-29)$$

Overall:

$$\dot{M}_{v_{i-1}} = \left[(N_A + N_B) M \right] \Delta A + \dot{M}_{vi} \quad (3-30)$$

For the liquid Phase:

Light component:

$$\dot{M}_{\ell_{i-1}} x_{\ell_{i-1}} + \left[(N_A + N_B) zM \right] \Delta A = \dot{M}_{\ell i} x_{\ell i} \quad (3-31)$$

Overall:

$$\dot{M}_{\ell_{i-1}} + \left[(N_A + N_B) M \right] \Delta A = \dot{M}_{\ell i} \quad (3-32)$$

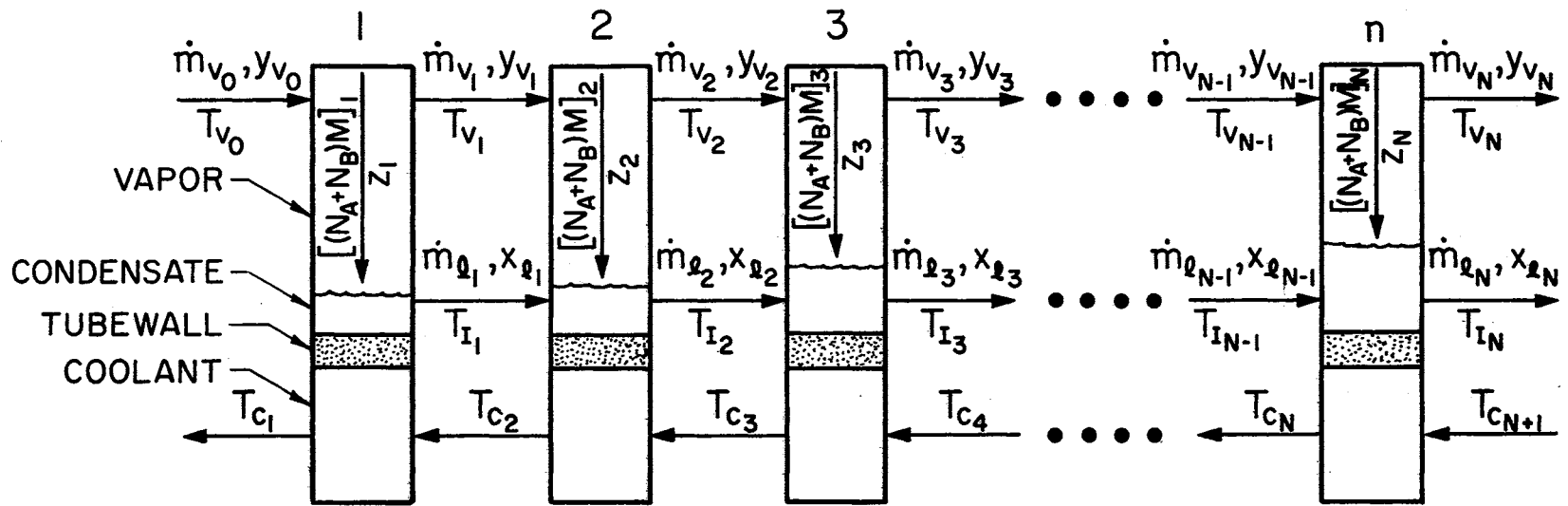


Figure 2. Schematic Diagram of Vapor, Condensate and Coolant Flows in the Condenser

Similarly an energy balance between the vapor and condensate and the coolant is written for each increment. The total heat load arises from three sources: the sensible and latent heat transfer from the vapor that is condensed in a given increment, the sensible cooling of the uncondensed vapor, and the sensible cooling of the condensate. The heat transferred from the condensing vapor is merely the heat flux given in Equation 3-9 times the area increment ΔA :

$$q_c = \left(\frac{Q}{A}\right)_T \Delta A \quad (3-33)$$

The sensible heat removed from the uncondensed vapor is:

$$q_{sv} = \dot{M}_v \bar{C}_{Pv} (T_{v_{i-1}} - T_{v_i}) \quad (3-34)$$

Where i corresponds to the increment in question and $i-1$ to the preceding increment. The sensible heat removed from the condensate is:

$$q_s = \dot{M}_c \bar{C}_{Pc} (T_{c_{i-1}} - T_{c_i}) \quad (3-35)$$

The sum of these three heat loads is the heat picked up by the coolant:

$$q_T = q_c + q_{sv} = \dot{M}_c \bar{C}_{Pc} (T_{c_i} - T_{c_{i+1}}) \quad (3-36)$$

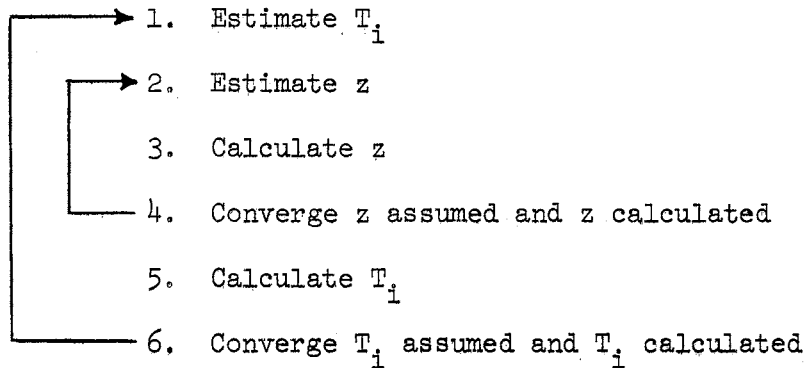
The temperature of the vapor leaving an area increment is calculated by solving this equation for T_{v_i} :

$$q_{sv} = (\bar{T}_v - T_i) h_G \Delta A = (T_{v_{i-1}} - T_{v_i}) \dot{M}_v \bar{C}_{Pv} \quad (3-37)$$

CHAPTER IV

THE COMPUTER PROGRAM

The main problem is the calculation of the interfacial temperature T_i and the composition z of the vapor condensed in an area increment. Since neither of these unknowns can be expressed as an explicit function of the independent system variables; a double trial and error calculation is required. In addition, if the enhanced mass transfer coefficient is used another smaller loop is required within the loop for z . The general scheme to converge to the correct values of T_i and z is:



The important point in this procedure is to converge the inner loop (z assumed \approx z calculated) at each iteration of the outer loop otherwise, the system may fail to converge. The convergence of z was chosen as the inner loop because, for the case using the unenhanced mass transfer coefficient, z may be found from the solution of two non-linear equations. On the other hand, T_i is a more complex function and was chosen for the outer loop. For the case where the "two phase"

mass transfer coefficient is used one additional equation is added to the inner loop for z . This increases the complexity of the convergence of z but the inner loop is still less complex. A Wegstein (19) type convergence method was used to converge both loops.

Once T_i and z are found, the next vapor temperature and composition and the individual heat loads may be calculated. The procedure is then repeated for all succeeding increments until the desired vapor temperature or total condensation is achieved. The overall calculation scheme is:

1. Choose an increment of area, ΔA .
2. Calculate physical properties at T_v .
3. Calculate F .
4. Assume T_i .
5. Calculate y_i and x_{li} by equilibrium calculation at T_i .
6. Assume z .
7. Calculate $(N_A + N_B)$ from liquid phase mass balance.
8. If the two phase enhancement of the mass transfer coefficient is used, calculate the quality x and the enhancement factor R .
9. Calculate z from diffusion equation.
10. Is z assumed = z calculated? If yes, go to 10; if no, go to 6.
11. Calculate h_c at T_i , y_i , P .
12. Calculate U .
13. Calculate h_G and perform heat flux balance on vapor phase.
14. Does the heat flux balance check? If yes, go to 14; if no, go to 4.

15. Calculate y_v for the next increment by mass balance on vapor phase.
16. Calculate T_v for the next increment by heat balance on q_{sl} .
17. Calculate q_{sv} , q_{sl} , q_λ and q_T .
18. Calculate T_w .
19. Continue until $T_v = T_v$ outlet, or total condensation has occurred.

The block flow diagram for the computer program may be found in Appendix C. The format for input data preparation is illustrated in Appendix B. The required input data includes: the inlet conditions of the vapor (temperature, pressure, composition and flow rate), data for the coolant stream, block physical property and equilibrium data, specified area increment, number of tubes, type of coefficient used (enhanced or single phase) and the orientation of the condenser.

The first step in the program is the calculation of the bubble point temperature from the input equilibrium data. This temperature TLTOT will be used in conjunction with the variable LTEST to check for the possibility of total condensation or an attempt by the computer to converge on an interfacial temperature below the bubble point of the feed.

The main loop is used to converge on the value of T_i for a given area increment. Near the beginning of the loop the variable LTEST is used. If the quality is less than .3 (arbitrarily set value), LTEST is set equal to one and the condensing stream will be checked for total condensation later in the program. Subroutines VAPOR and DCOEF are used to obtain the vapor physical properties at the vapor bulk

temperature, pressure and composition. Once the physical properties are known, the Reynolds number, Prandtl number, Schmidt number, and mass transfer coefficient are calculated.

Once T_i is assumed, LTEST is checked. If LTEST is zero, the program continues in the main loop. If LTEST is one, the condensing stream is tested for total condensation by the use of subroutines EQUIL and TTEST. Subroutine EQUIL is used to calculate the equilibrium compositions at T_i (for this case it is the bubble point TLTOT). TTEST performs the Colburn-Drew calculations in slightly different form. In this case the area increment is calculated by knowing the interfacial temperature and the condensing rate ($N_A + N_B$). If this calculated area is less than the specified area, the stream will reach its bubble point with less area than is specified. In this event, the program leaves the main loop and the heat loads are calculated for the final area increment. If the calculated area is greater than the specified area increment, then LTEST is set equal to zero and the program continues in the main loop.

The value of z is now required for the calculations and is calculated in a slightly different manner depending on whether the single phase heat and mass transfer coefficients are used or the enhanced coefficients are used. If the unenhanced version of the coefficients is used then the value of z is obtained from the simultaneous solution of Equations (3-5) and (3-31) utilizing subroutine CONV. An exception to this is for the first area increment where z is equal to x_{1i} and the loop is skipped entirely. If the enhanced coefficient is used, an additional sub-loop is required because the total flux ($N_A + N_B$) is directly proportional to the mass transfer coefficient; which in turn

is related to $(N_A + N_B)$ through the quality in the Martinelli-Nelson factor. The procedure used was to converge on a value of F by using an interval halving technique each time through the z loop. If the program is in the first area increment the value of z is known and only the sub-loop for F is used.

Once the loop for z is converged to an acceptable value, the condensing coefficient and overall coefficient are calculated. Subroutine CALC is then used to perform the Colburn-Drew calculations. If the heat flux equation balances within a specified tolerance or the calculated interfacial temperature is close enough to the assumed value, the results are printed and heat loads, next vapor temperature, and next vapor composition are calculated. The program then returns to the beginning of the loop for the next area increment calculation. If the heat balance or calculated temperature do not check the loop is continued until convergence is achieved. This routine is applied for all area increments until the desired temperature or the bubble point temperature is reached.

CHAPTER V

RESULTS AND CONCLUSIONS

In order to demonstrate the operation of the computer program developed, two example problems were solved. The results are compared with those of the approximation method formulated by Bell and Ghaly (12). The first problem was concerned with a methanol-water system similar to that used by Colburn and Drew (1). The other problem was a system of n-butane and n-octane chosen because of the wide condensing temperature range. Both problems assumed condensation inside a vertical bundle of 100-1 inch 14 BWG tubes (.834 in I.D.). Water was used as the cooling medium with a corresponding heat transfer coefficient of 900 Btu/hr ft² °F. The fouling coefficient for each stream was assumed to be 1000 Btu/ft²hr°F. The conditions used for the methanol-water system were:

Mass flow rate: 5000 lb/hr

Composition: .30 mole fraction methanol

Inlet temperature: 197.0°F (Saturated vapor)

Pressure: 1 atmosphere absolute

Water conditions: 10⁵ lb/hr at 68°F

The conditions assumed for the n-butane - n-octane system were:

Mass flow rate: 8000 lb/hr

Composition: .30 mole fraction n-butane (Saturated vapor)

Temperature: 259.1°F

Pressure: 20 psia

Water conditions: 5×10^4 lb/hr at 86°F

The temperature profiles for the methanol-water system are shown in Figure 3. Both the vapor bulk and interfacial temperatures are plotted as a function of heat transfer area for the Colburn-Drew method with and without two phase enhancement of the vapor phase coefficients and for the approximate method. The curves indicate that with the two phase enhancement, the vapor bulk temperature is less superheated with respect to the interface than the corresponding vapor bulk temperature for the unenhanced case.

The two phase heat transfer coefficient tends to keep the vapor bulk temperature down while the enhanced mass transfer coefficient tends to keep the interfacial temperature higher. This latter statement may be supported by referring to Equations (3-9) and 3-10).

Since F and h_G are quite larger, the temperature difference ($T_i - T_w$) in Equation (3-10) must be large to balance with Equation (3-9). In either case, the temperature profiles indicate that there is only a relatively minor resistance to heat transfer in the vapor phase since the greatest temperature difference (about 100°F) is between the interface and the coolant.

Figure 4 shows a plot of the vapor bulk composition as a function of area. The Colburn-Drew method without two phase enhancement predicts a lower concentration increase of the light component than does the approximation method. This follows from the temperature profiles in Figure 4. The interfacial temperature for the unenhanced method is lower than the bulk temperature of the approximate method. Since the composition of the liquid phase increases in decreasing temperature,

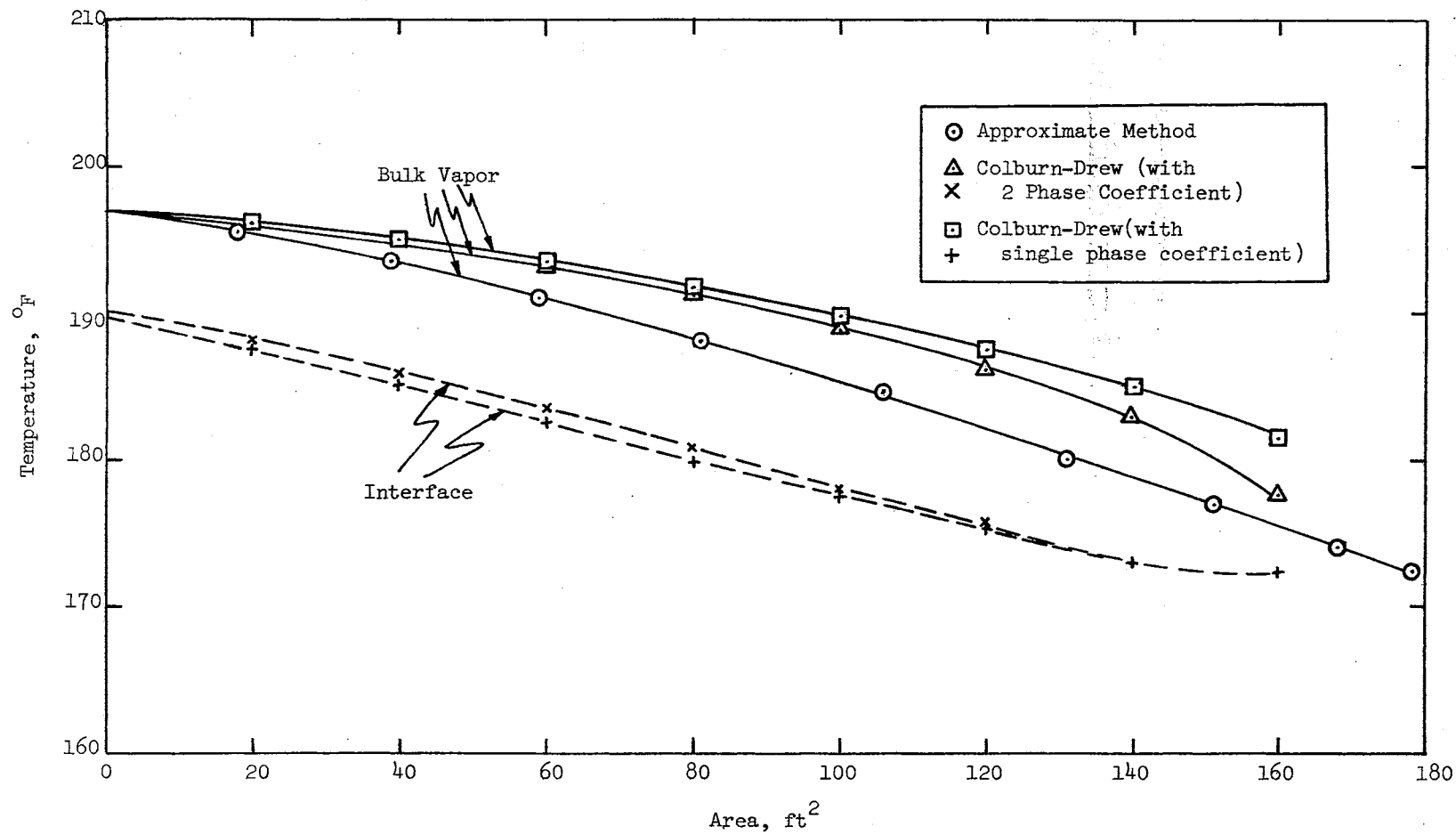


Figure 3. Temperature Profiles for the Methanol-Water System

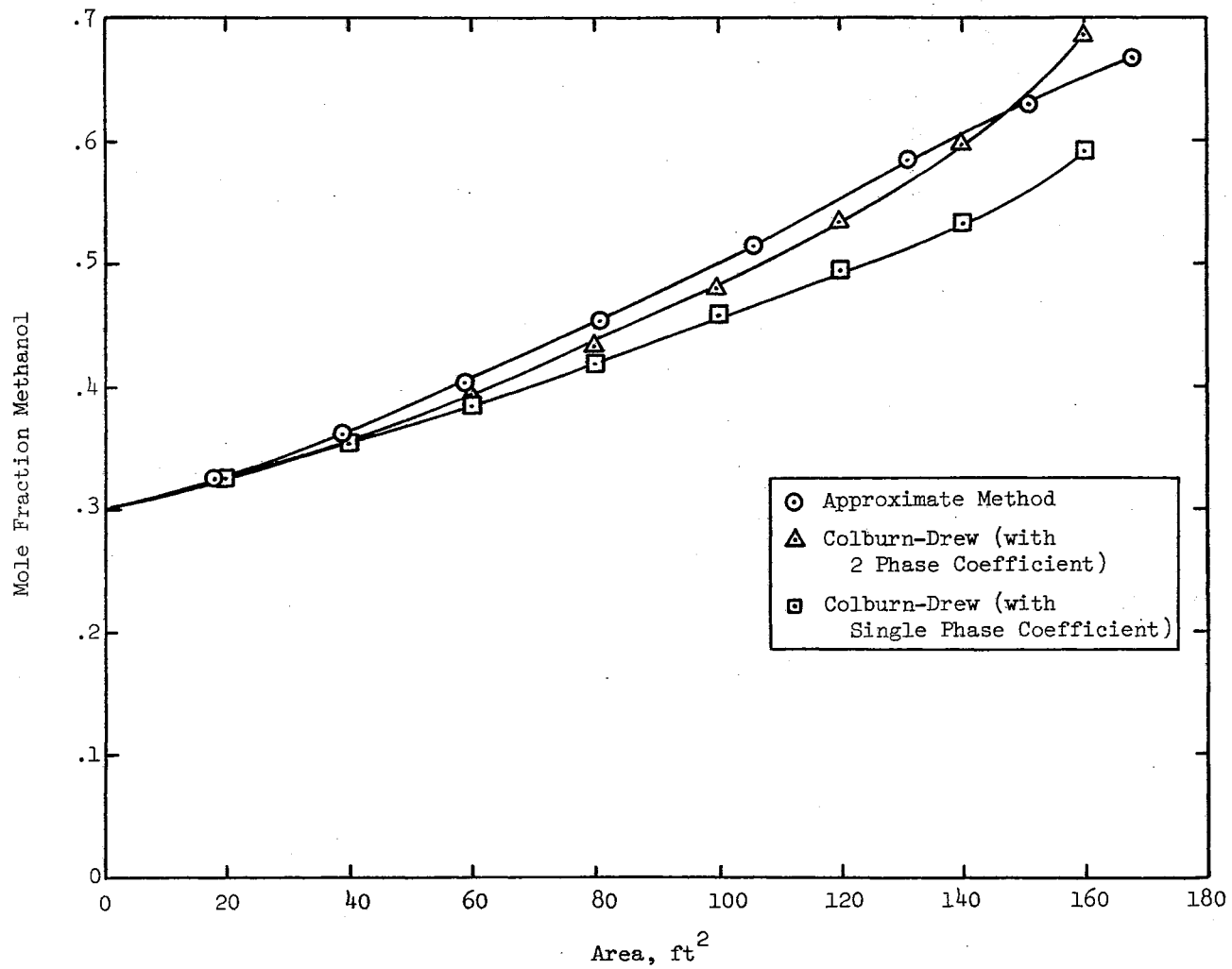


Figure 4. Bulk Vapor Composition Vs. Area for the Methanol-Water System

the composition of the liquid in the Colburn-Drew case would be higher than that of the approximate method. By an overall material balance at any point the composition of the vapor must be less for the Colburn-Drew method with no enhancement than for the approximate method. Another factor which enters in is the build up of the light component in the vapor phase at the interface which tends to decrease the composition of the bulk vapor. The method using the enhanced coefficients showed a lower concentration increase than the approximate method until the end of the process where the enhancement factor for the mass transfer coefficient is significantly greater than one. The compositions near the end of each process are nearly identical, indicating that if the two methods had more nearly equal condensing rates, the compositions would not be too different.

Figure 5 indicates that the approximate method, while being somewhat conservative, predicts an area which is less than 20% greater than the Colburn-Drew methods. This is to be expected since, as we pointed out before, the major resistances for this case appear to be the coolant, condensate, and fouling resistances which are the same for all methods.

The vapor temperature profiles for the n-butane-n-octane problem are shown in Figure 6. Since the difference in the dew point and bubble point for this mixture is about 140°F the effect of the sensible heat transfer rate in the vapor phase is much more pronounced. Also, the vapor phase transport processes are much more important in limiting the heat transfer rate. The temperature difference between the bulk and interfacial temperatures for the enhanced case is about 40°F ;

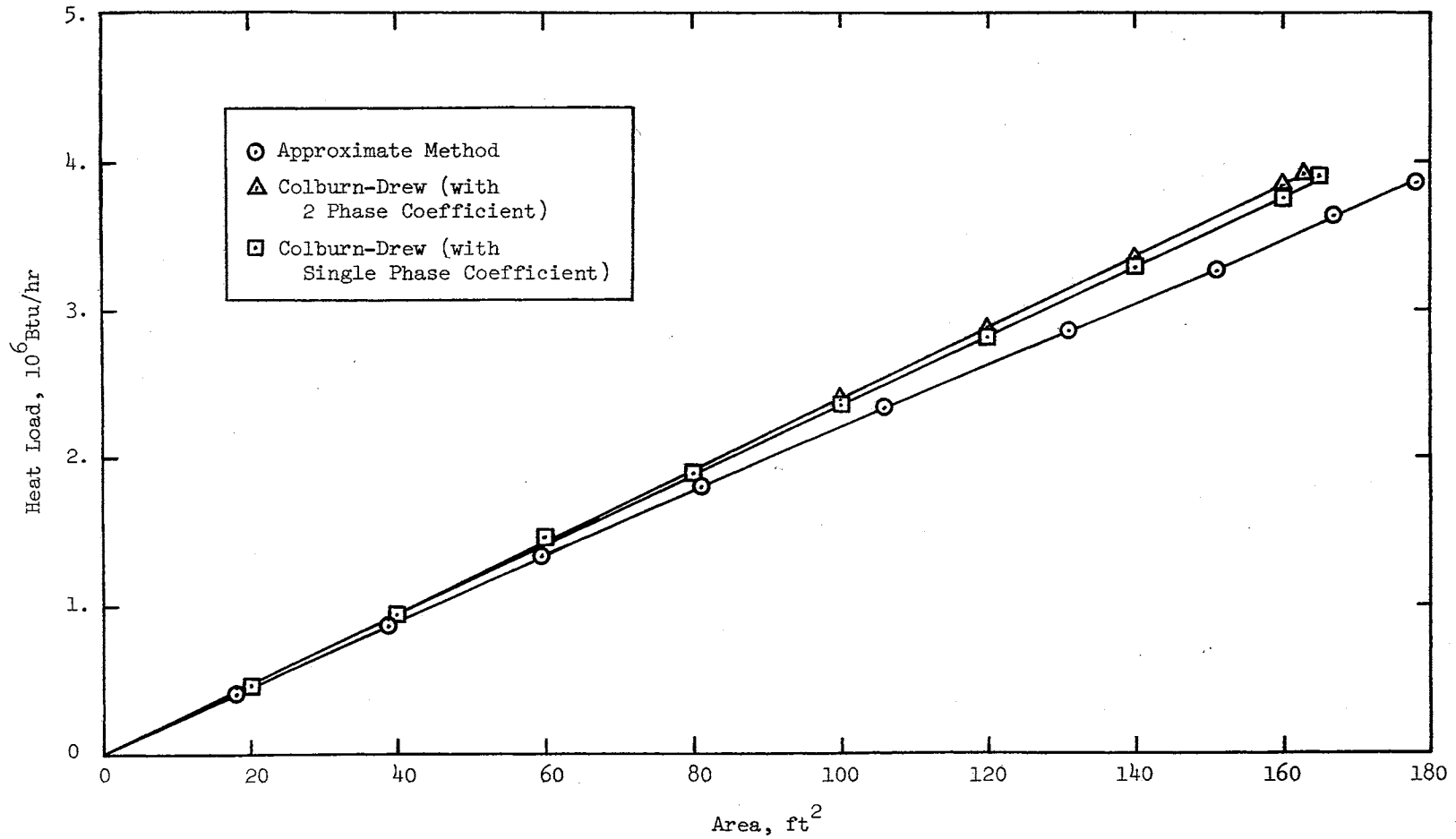


Figure 5. Heat Transferred to the Coolant for the Methanol-Water System

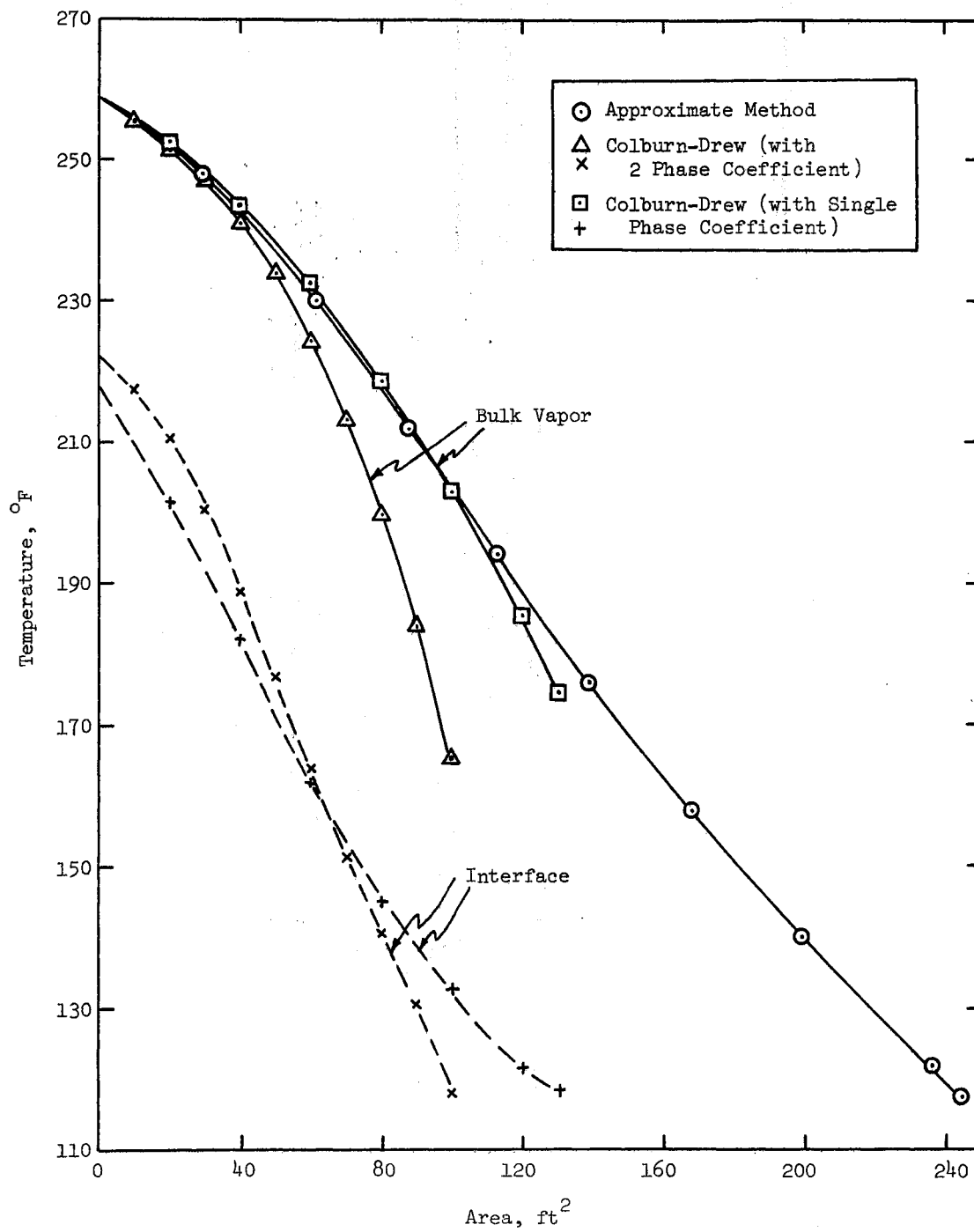


Figure 6. Temperature Profiles for the n-Butane-n-Octane System

whereas, for the unenhanced case the difference is 50-60°F. This corresponds to a much greater resistance in the vapor phase than was exhibited in the methanol-water case.

The composition profiles in Figure 7 are similar qualitatively to those for the methanol-water case, but the differences between the profiles are much more exaggerated because of the increased vapor phase resistances.

The heat load profiles in Figure 8 show that for this case the approximation method overestimates the required heat transfer area. This is due largely to the overcompensation for the sensible heat load in the vapor phase.

The cases presented here seem to be two limiting cases. For the methanol-water system the vapor phase resistances to heat and mass transfer are quite small; whereas, for the n-butane-n-octane system the vapor phase resistances are large and must be accounted for in the condenser design.

The results of these two cases indicate that the more fundamental Colburn-Drew approach to condenser design leads to less conservative designs, further study is needed to really determine whether the Colburn-Drew method gives a "true" value of the area required. The approximate method is conservative in the predicted heat transfer area. This is in the direction which is desirable; but, this conservatism can be costly in an actual design application if it is used for a case such as the n-butane-n-octane system. On the other hand the Colburn-Drew method may be too non-conservative and a condenser may be designed which will not perform the desired condensation.

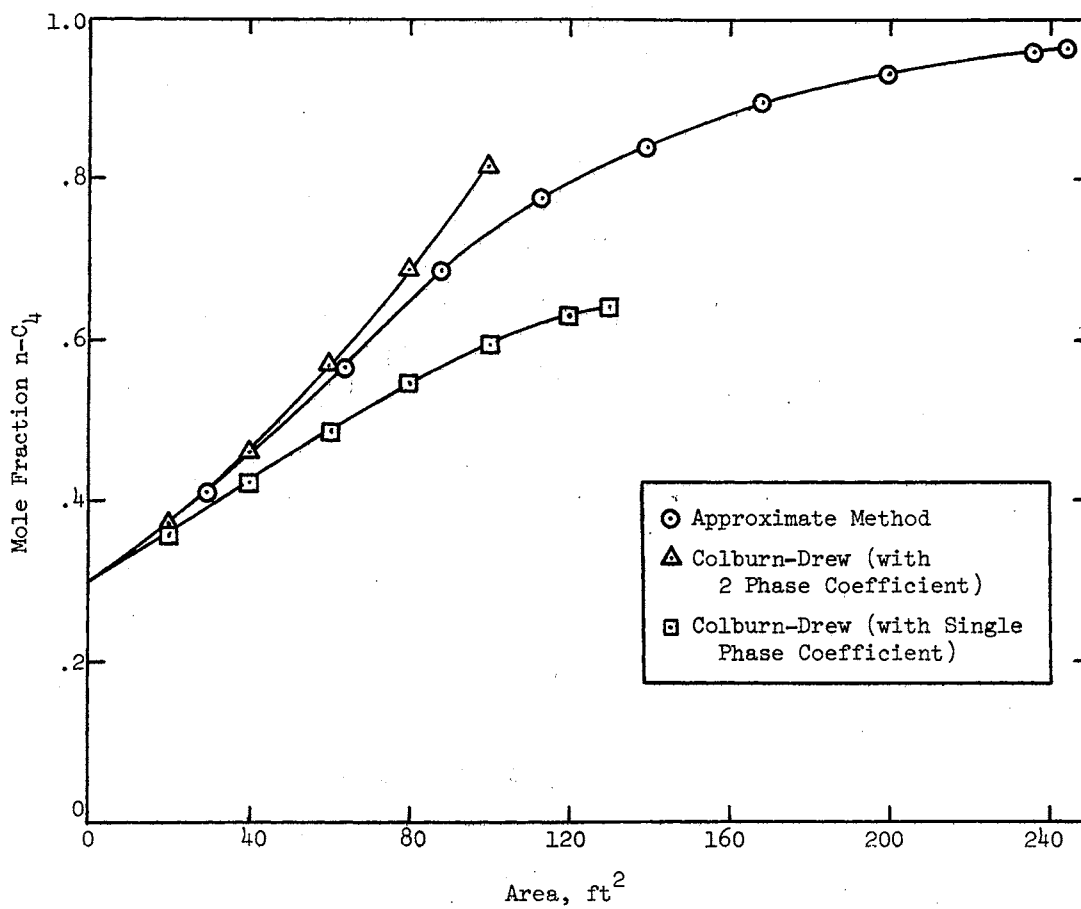


Figure 7. Vapor Composition Vs. Area for the n-Butane-n-Octane System

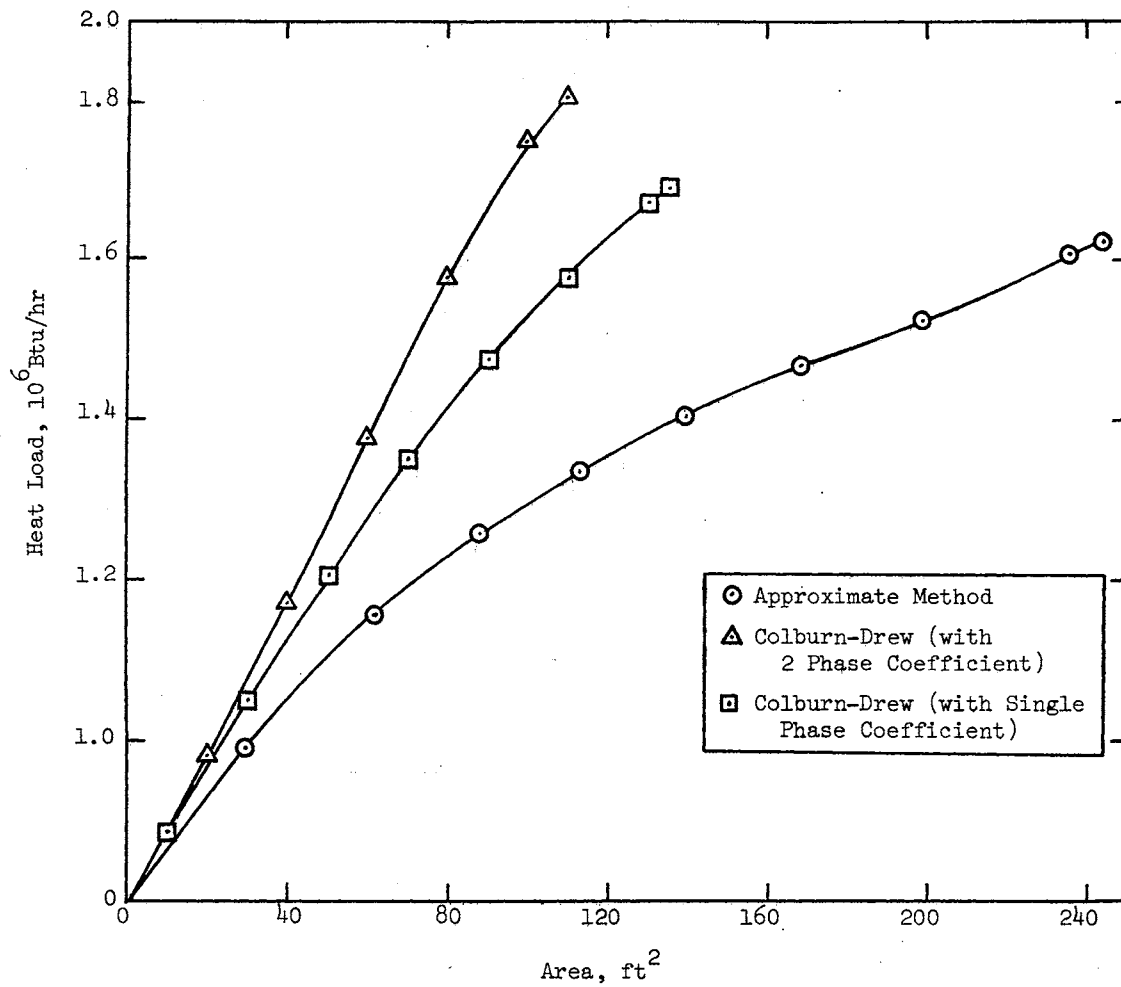


Figure 8. Heat Transferred to the Coolant for the n-Butane-n-Octane System

The proposed method, while being limited to binary vapors, can probably be extended to multicomponent mixtures. This would require further sophistications of the subroutines for the thermodynamics and physical properties associated with multicomponent systems.

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APPENDIX A
COMPUTER PROGRAM FOR THE SOLUTION
OF THE COLBURN-DREW EQUATIONS

COMPUTER PROGRAM FOR THE SOLUTION
OF THE COLBURN-DREW EQUATIONS

The computer program for the design of binary vapor condensers using the Colburn-Drew equations is written in FORTRAN IV for use on the IBM 360 model 65 digital computer. A block diagram of the main program is shown in Appendix B. A description of the main program and the subroutines is presented in the following sections. Appendix C presents selected values of the main variables for each of the two systems studied.

Main Program

This is the executive program for the entire calculational scheme. The subroutines for the thermodynamic and physical properties and the basic calculations are called by the main program. The basic logic for the main program is discussed in Chapter III.

Subroutine VAPOR

This subroutine supplies the main program with the necessary physical properties for the vapor phase. The vapor density is calculated from the ideal gas law corrected with the compressibility factor Z_{RK} :

$$\rho_v = \frac{PM}{T_v Z_{RK}} (10.73) \quad (A-1)$$

The compressibility factor is obtained by the solution of the Redlich-Kwong equation in subroutine RK.

The pure component heat capacities are found in the equation:

$$C_{Pi} = A + BT + CT^2 \quad (A-2)$$

where A, B, and C are constants supplied from input data. The heat capacity of the vapor is then found from a mole fraction average of the pure component heat capacities.

The thermal conductivity is calculated by a mole fraction average of the pure component thermal conductivities. The pure component values are found from a linear interpolation of the input data by use of function FUN1.

The vapor phase viscosity is computed by using a correlation by Wilke (20):

$$\mu_v = \frac{\mu_1}{\left(1 + \left(\frac{y_2}{y_1}\right)\phi_{12}\right)} + \frac{\mu_2}{\left(1 + \left(\frac{y_1}{y_2}\right)\phi_{21}\right)} \quad (A-3)$$

where:

$$\phi_{12} = \frac{\left[1 + \left(\frac{\mu_1}{\mu_2}\right)^{1/2} \left(\frac{M_1}{M_2}\right)^{1/4}\right]^2}{\left(\sqrt{8} + 1 + \left(\frac{M_1}{M_2}\right)\right)^{1/2}} \quad (A-4)$$

and:

$$\phi_{21} = \left(\frac{\mu_2}{\mu_1}\right) \left(\frac{M_1}{M_2}\right) \phi_{12} \quad (\text{A-5})$$

The pure component viscosities μ_1 and μ_2 are found by using function FUN1.

Subroutine LIQ

This subroutine is used to supply the liquid phase physical properties required. The heat capacity, thermal conductivity, and viscosity are all found from a mole fraction average of the pure component data obtained by using function FUN1. The density is also found from a linear interpolation of the input data for saturated liquid densities.

Subroutine DCOEF

The diffusion coefficient is calculated in this subroutine by a method given in Reid and Sherwood (20):

$$b = .001858 T_v^{3/2} \left(\frac{M_1 + M_2}{M_1 M_2}\right)^{1/2} / P \sigma_{12}^2 \Omega_D \quad (\text{A-6})$$

where σ_{12}^2 and Ω_D are empirical constants.

Subroutine COEF

The calculation of the condensing coefficient is performed using various methods depending on which is applicable. For the vertical orientation, Nusselt's equation, a fit of Colburn's correlation, and the Boyko-Kruzhilin method are all calculated and the highest value

selected since this is the one which is valid for the particular case. For horizontal orientation, the Kern correlation and the Boyko-Kruzhilin correlation are used. A discussion of all these equations is presented in Chapter III.

Subroutine CALC

This subroutine performs the Colburn-Drew calculations as outlined in Chapter IV. The main purpose of these calculations is to obtain both sides of the heat flux balance for the vapor phase. Once these calculations are performed, the computer returns to the main program to check for convergence in the heat flux balance.

Subroutine WRITE

This subroutine is used to print the results of the calculations performed at the end of each area increment and at the end of the entire program. Various sections of the subroutine are executed according to the value of LPRT which is set in the main program before the subroutine is called.

Subroutine TTEST

This subroutine is used to check for total condensation within an area increment. The condensation rate and mole fraction are known. The area required for total condensation is calculated. If the calculated area is less than the area increment, TTEST is set equal to zero and the computer returns to the main loop in the main program.

Functions FUN1 and FUN3

These functions are used for the linear interpolation of the thermodynamic and physical property data. The difference in the functions is FUN1 is for increasing values of the independent variable and FUN3 is for decreasing values.

Subroutine EQUIL

This subroutine computes the required equilibrium mole fractions at any temperature. These compositions are obtained by linear interpolation of the T-x-y data using FUN1.

Subroutine CONV

This subroutine is shown in a block diagram in Appendix B. The purpose of this subroutine is to converge the value of Z in the main program by using Wegstein's method (19): The main idea of the procedure is shown in Figure 9. The basic idea is to project from two known points on the $f(X)$ line to the $X=X_c$ line to determine the new trial value. In order to project to the diagonal line, this equation must be solved for the new trial value X_3 :

$$X_3 = \frac{X_1 X_{c2} - X_{c1} (X_2)}{X_1 - X_2 + X_{c2} - X_{c1}} \quad (\text{A-7})$$

Once this new trial value is found it is substituted into $f(X)$ to determine the new calculated value. If these values are close enough

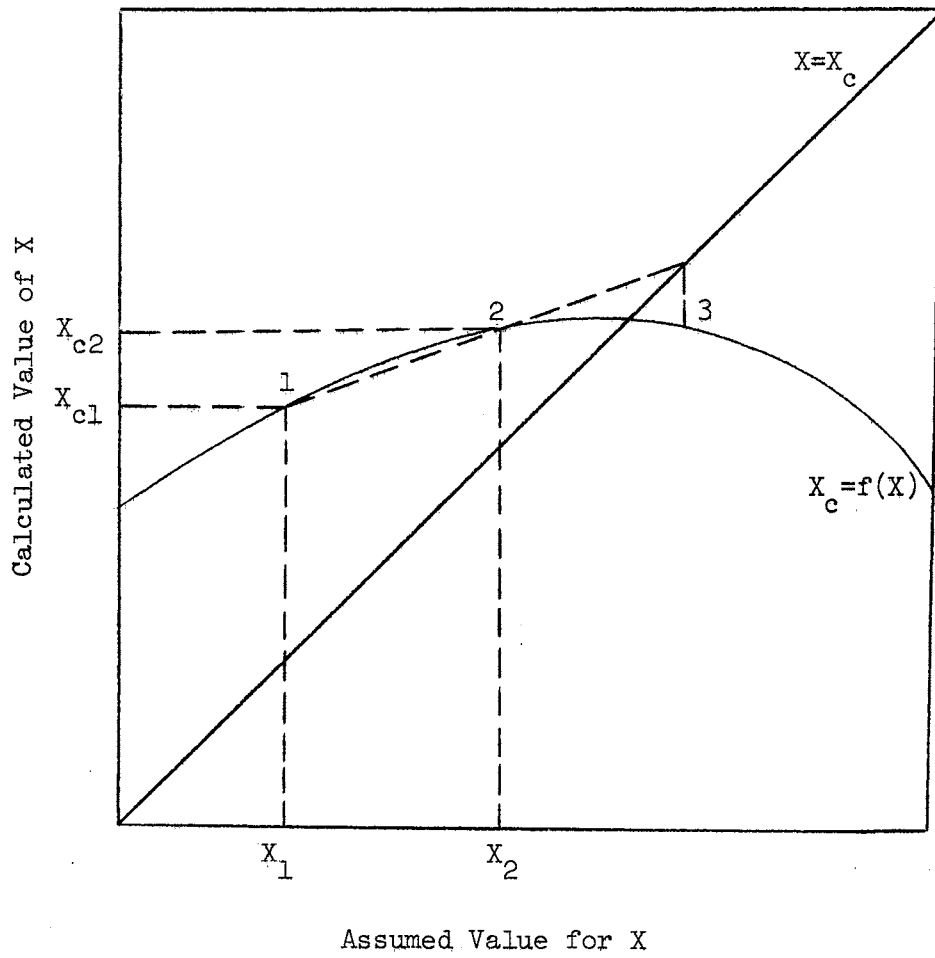


Figure 9. Wegstein Convergence Method.

for convergence, the variable NC is set equal to one. If the values are not close enough, NC is two and the procedure is repeated until adequate convergence is achieved.

Subroutine CONV

This subroutine is a modified form of subroutine CONV. Instead of using assumed and calculated values of a variable for convergence, the assumed interfacial temperature and difference in the heat balance is used. The difference is projected to zero and the new assumed interfacial temperature is found by this projection.

Subroutine RK

This subroutine solves the Redlich-Kwong equation of state (21) to determine the compressibility factor used in subroutine VAPOR. The form of the equation is:

$$Z_{RK} = \frac{1}{1-h} - \frac{A^2}{B} \left(\frac{h}{1+h} \right) \quad (A-8)$$

where:

$$A^2 = \frac{a}{R^2 T^{5/2}}; \quad a = \frac{.42748 R^2 T_c^{5/2}}{P_c} \quad (A-9)$$

$$B = \frac{b}{RT}; \quad b = \frac{.08664035 RT_c}{P_c}; \quad h = \frac{BP}{Z}$$

The problem is the solution of the cubic equation which results.

Subroutine APPR

This subroutine is for the condenser design using the approximate method by Bell and Ghaly (12). The procedure is to calculate the function described in Equation 2-1 and graphically integrate the function versus Q .

Subroutine SUBCL

This subroutine provides for subcooling of the condensed stream. The required heat load is calculated and the area required to perform the subcooling is calculated and added to the area required for condensation.

Data Preparation

The input data cards required for this program are arranged in the following order:

CARD 1

This card contains data for four variables. The input variables are:

TYPE = value determines design method(s) used. If it is 1, the Colburn-Drew is used; if 2, both the Colburn-Drew and the approximate methods are used; if 3, only the approximate method is used.

NPHAZ = this picks the type of heat and mass transfer coefficient used for the vapor phase. If the value is 1, the unenhanced coefficients are used; if the value is 2, the enhanced coefficients are used.

ORIENT = type of orientation desired. for vertical case, use a 1; for the horizontal case it should be 2.

TUBES = number of tubes in the condenser.

READ: TYPE, NPHAZ, ORIENT, TUBES

FORMAT: 3I2, F6.1

CARD 2

This card supplies data for the condensing stream.

TV1 = inlet temperature of vapor, °F

TV2 = outlet temperature of the stream, °F

MDOT = total inlet flow rate of vapor, lb/hr

SUMR = sum of the heat transfer resistances (except for the condensate film) for the calculation of the overall coefficient.

Y1 = inlet vapor composition (mole fraction of the light component).

MA = molecular weight of light component, lb/lb mole.

MB = molecular weight of heavy component, lb/lb mole.

READ: TV1, TV2, MDOT, SUMR, Y1, MA, MB

FORMAT: 7E10.5

CARD 3

This card contains data for the water stream and additional data for the condensing stream.

TW1 = water outlet temperature, °F

FLOW = water flow rate, lb/hr

PROB = problem number (for identification purposes only)

NT = maximum number of area increments allowed (can be as large as 49)

D = tube inside diameter, inches

DELA = area increment size, ft²

P = pressure for condensing stream, psia

CARDS 4 thru N

The number of cards in this section is dependent upon the user. The order of the cards is quite rigid and failure to adhere to the format will result in a program which will not run.

Section 1: vapor thermal conductivity data. The input variables are:

TKV = temperatures at which the data points are provided,
 $^{\circ}\text{F}$

XKV1 = thermal conductivity of the heavy component BTU/
 $\text{ft hr}^{\circ}\text{F}$

XKV2 = thermal conductivity of the heavy component BTU/
 $\text{ft hr}^{\circ}\text{F}$

(Data must be read in with increasing temperature.)

READ: TKV(I), XKV1(I), XKV2(I)

FORMAT: 3E10.4

NOTE: A blank card must follow the last data card for this section.

SECTION 2: liquid thermal conductivity data. The input variables are:

TKL = temperatures at which the data points are provided,
 $^{\circ}\text{F}$

XKL1 = thermal conductivity of the light component,
 $\text{BTU}/\text{ft hr}^{\circ}\text{F}$

XKL2 = thermal conductivity of the heavy component,
 $\text{BTU}/\text{ft hr}^{\circ}\text{F}$

(Data must be read in with increasing temperature.)

READ: TKL(I), XKL1(I), XKL2(I)

FORMAT: 3510.4

NOTE: A blank card must follow the last data card for this section.

Section 3: liquid density, latent heats, and equilibrium compositions data. The input variables are:

T = the temperature at which the data points are provided

XLDEN = liquid density data, lb/ft³

XLAM1 = latent heat of the light component, BTU/lb mole

XLAM2 = latent heat of the heavy component, BTU/lb mole

X = liquid mole fraction of the light component

Y = vapor mole fraction of the light component

NOTE: A blank card must follow the last data card for this section.

Section 4: liquid heat capacity data. Input variables are:

TCP = temperatures at which data points are supplied

CPL1 = heat capacity of the light component, BTU/lb^oF

CPL2 = heat capacity of the heavy component, BTU/lb^oF

(Data must be read in order of increasing temperature.)

READ: TCP(I), CPL1(I), CPL2(I)

FORMAT: 3E10.4

NOTE: A blank card must follow the last data card for this section.

Section 5: constants for the vapor heat capacity equations.

Input variables are:

ACP,BCP,CCP = constants for the equation of the form:

$$ACP + BCP \times 10^{-3} T + CCP \times 10^{-6} T^2$$

READ: ACP(I), BCP(I), CCP(I)

FORMAT: 3E10.4

Constants for the light component are read from the first card and constants for the heavy component on the second card.

Section 7: vapor viscosity data. Input variables are:

TMU = temperatures at which the data points are provided

XMU1 = viscosity of the light component, cp.

XMU2 = viscosity of the heavy component, cp.

(Data must be read in order of increasing temperature.)

NOTE: A blank card must follow the last data card for this section.

Section 8: liquid viscosity data. Input variables are:

TYMU = temperatures at which the data points are provided

YMU1 = viscosity of the light component, cp.

YMU2 = viscosity of the heavy component, cp.

(Data must be read in order of increasing temperature.)

NOTE: A blank card must follow the last data card for this section.

CARD N+1 Data for the diffusion coefficients is read in:

T1,T2 = Leonard-Jones parameters σ_1 , and σ_2 are tabulated in Reid and Sherwood (20), Å

EK1,EK2 = Leonard-Jones parameters (ϵ/K), and (ϵ/K) as tabulated in Reid and Sherwood (20), °K

CARD N+2 to (N + NT) + 2

The input data on the next NT cards (where NT is the number of steps read in on card 1) contain the vapor temperatures used in the approximate method.

TV = vapor bulk temperature, °F

READ: TV(I)

FORMAT: E10.4

It is very important that the exact number of steps NT correspond with the number of temperatures when the approximate method is used.

APPENDIX B
FLOW DIAGRAMS

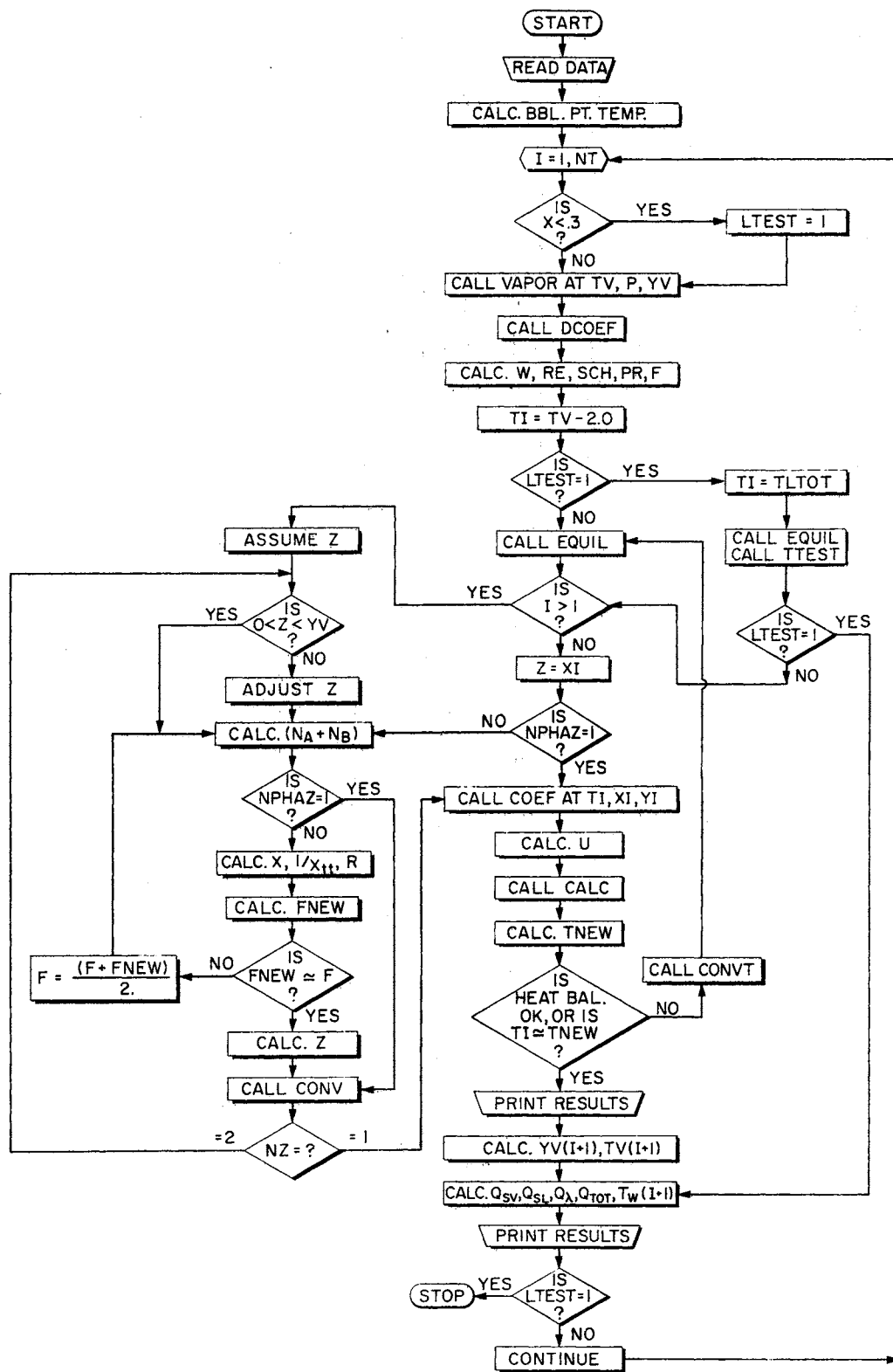


Figure 10. Flow Diagram for the Main Program

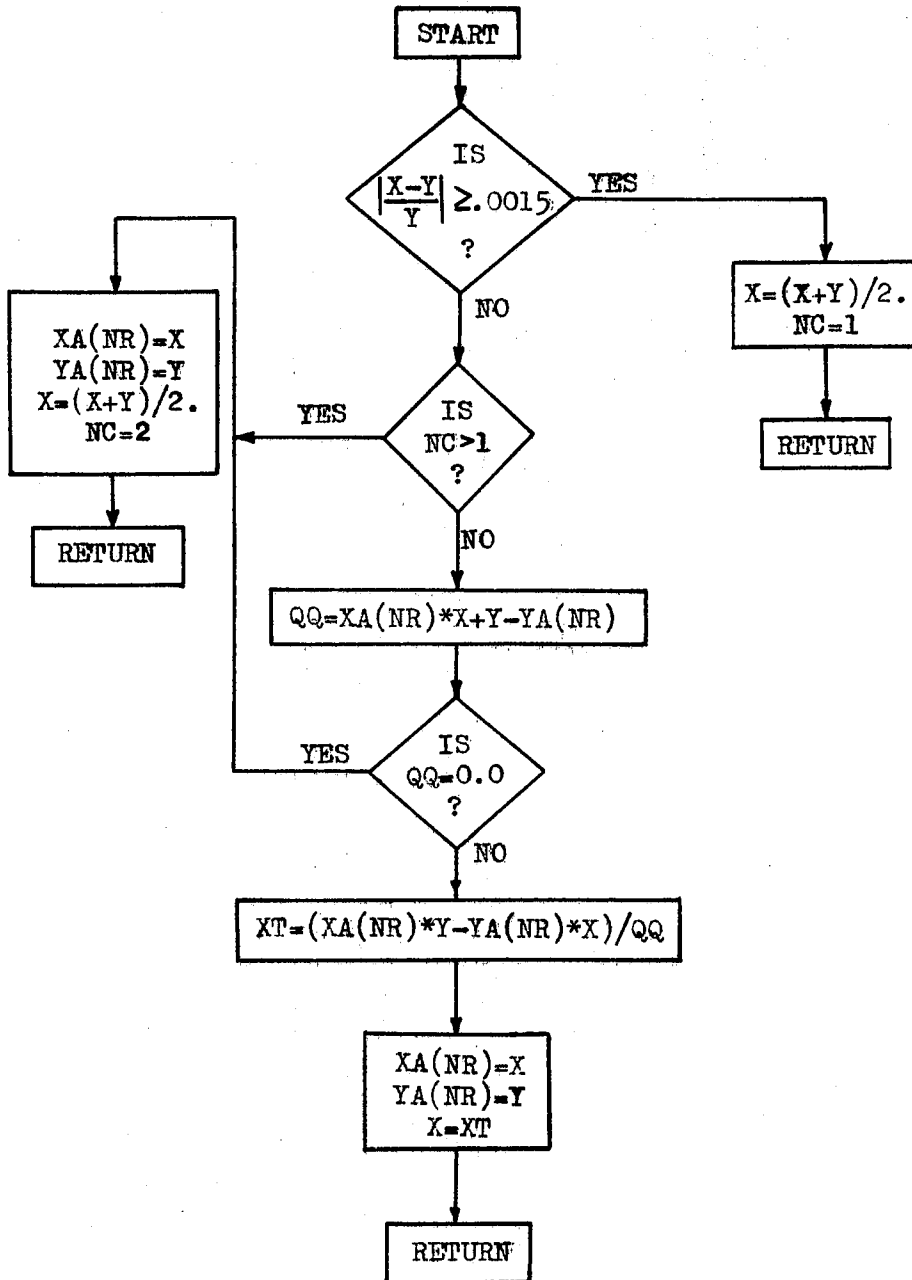


Figure 11. Flow Diagram for CONV

APPENDIX C
TABULATED RESULTS

TABLE I
RESULTS OF THE COLBURN-DREW CALCULATIONS WITHOUT
ENHANCEMENT FOR THE METHANOL-WATER SYSTEM

Vapor Bulk Temp, °F	Interfacial Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
197.0	189.7	104.0	.300	5000	0	0
196.2	187.5	99.4	.326	4490	.463	20
194.9	185.1	94.5	.355	3956	.934	40
193.5	182.5	89.9	.386	3396	1.410	60
191.8	179.8	85.1	.420	2813	1.887	80
189.7	177.5	80.4	.458	2210	2.359	100
187.5	175.4	75.8	.496	1581	2.823	120
185.0	173.1	71.2	.534	936	3.278	140
181.6	172.4	66.5	.592	209	3.751	160
172.4	---	65.2	---	0	3.881	165

TABLE II
 RESULTS OF THE COLBURN-DREW CALCULATIONS WITH
 ENHANCEMENT FOR THE METHANOL-WATER SYSTEM

Vapor Bulk Temp, °F	Inter-facial Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
197.0	190.2	104.0	.300	5000	0	0
196.0	188.3	99.3	.327	4488	.467	20
194.8	186.1	94.6	.358	3952	.944	40
193.2	183.7	89.7	.393	3391	1.429	60
191.4	181.0	84.8	.433	2805	1.917	80
189.1	178.3	80.0	.481	2196	2.405	100
186.3	175.9	75.2	.535	1564	2.883	120
183.0	173.2	70.5	.597	893	3.353	140
177.7	172.4	65.6	.687	144	3.842	160
172.4	---	64.8	---	---	3.923	163

TABLE III
 RESULTS OF THE APPROXIMATE METHOD
 FOR THE METHANOL-WATER SYSTEM

Vapor Bulk Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
197.0	104.0	.300	5000	0	0
195.5	99.9	.326	4571	.205	18
193.5	95.2	.362	4065	.871	39
191.1	90.7	.403	3554	1.329	59
188.1	86.0	.453	2998	1.803	81
184.5	80.7	.514	2341	2.327	106
180.1	75.6	.584	1645	2.843	131
176.9	71.5	.631	1024	3.254	151
174.2	67.7	.668	213	3.625	168
172.4	65.5	--	0	3.849	178

TABLE IV
 RESULTS OF THE COLBURN-DREW METHOD
 WITHOUT ENHANCEMENT FOR THE
 n-BUTANE-n-OCTANE SYSTEM

Vapor Bulk Temp, °F	Inter-facial Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
259.1	217.6	122.0	.300	8000	0	0
256.0	209.8	118.6	.329	7007	.172	10
252.4	201.5	115.2	.359	6108	.340	20
248.3	192.2	112.0	.390	5303	.499	30
243.6	182.1	108.9	.423	4573	.654	40
238.3	171.9	105.8	.455	3893	.808	50
232.4	161.9	102.9	.486	3271	.956	60
225.8	152.9	100.0	.517	2712	1.097	70
218.7	145.2	97.4	.546	2208	1.228	80
211.2	138.9	95.0	.573	1746	1.347	90
203.3	132.7	92.9	.596	1312	1.456	100
194.8	126.8	90.8	.616	910	1.558	110
185.6	121.5	88.9	.632	540	1.653	120
174.7	118.5	87.2	.643	198	1.738	130
118.5	---	86.3	--	0	1.786	135

TABLE V
 RESULTS OF THE COLBURN-DREW METHOD
 WITH ENHANCEMENT FOR THE
 n-BUTANE-n-OCTANE SYSTEM

Vapor Bulk Temp, °F	Inter- facial Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
259.1	222.4	122.0	.3000	8000	0	0
255.9	217.4	118.3	.3320	6922	.183	10
252.0	210.7	114.7	.3693	5910	.367	20
247.3	200.6	111.0	.4118	4986	.550	30
241.3	188.8	107.0	.4594	4118	.748	40
233.8	176.8	102.9	.5126	3321	.954	50
224.6	163.9	98.8	.5679	2595	1.161	60
213.3	151.4	94.6	.6267	1957	1.368	70
199.9	140.8	90.7	.6879	1398	1.563	80
184.3	130.7	87.2	.7492	893	1.729	90
165.3	118.2	84.0	.8183	440	1.902	100
118.2	---	80.5	---	--	2.077	111

TABLE VI
 RESULTS OF THE APPROXIMATE METHOD
 OF THE n-BUTANE-n-OCTANE SYSTEM

Vapor Bulk Temp, °F	Water Temp, °F	Vapor Bulk Comp.	Vapor Flow lb/hr	Heat Load, 10 ⁶ BTU/hr	Area ft ²
259.1	122.0	.300	8000	0	0
248.0	114.4	.412	5263	.382	29
230.0	107.8	.564	3279	.711	62
212.0	103.7	.684	2301	.915	88
194.0	100.6	.777	1709	1.072	113
176.0	97.9	.839	1293	1.207	140
158.0	95.3	.896	936	1.336	168
140.0	92.6	.931	552	1.470	199
122.0	89.7	.957	98	1.615	236
118.5	89.0	--	0	1.646	244

VITA

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Master of Science

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